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(54) Improvements in fluidized bed polymerization reactors.

(57) A fluidized bed reactor and operating process is dis-
closed, the reactor including distributor plate means to pro-
vide distribution of fluid passing upwardly into a fluidized bed;
a mixing chamber positioned below the distributor plate
means; fluid entry means, preferably at the base of the reactor,
for the passage of fluid into the reactor; flow deflector means
associated with the entry means for providing at least an up-
wardly fluid flow path and an outward fluid flow path along the
wall of the mixing chamber to inhibit the accumulation of solids
and liquid in the mixing chamber.

IMPROVEMENTS IN FLUIDIZED BED POLYMERIZATION REACTORS

The present invention relates to fluidized bed polymerization reactors. More particularly, it relates to improvements in fluidized bed polymerization reactors and the distribution of fluid introduced into such reactors.

The discovery of the fluidized bed process for the production of polymers provided a means for producing a diversity of polymers, e.g., polyolefins such as polyethylene, with a drastic reduction in capital investment and a dramatic reduction in energy requirements as compared to the conventional processes. However, a limiting factor in the use of a fluidized bed reactor for carrying out an exothermic polymerization process is the rate at which heat can be removed from the bed.

The most common and perhaps universal means of heat removal employed in conventional fluidized bed reactor processes is by compression and cooling of the recycle gas stream at a point external to the reactor. In commercial scale fluidized bed reaction systems for producing polymers such as polyethylene, the amount of fluid which can be circulated to remove the heat of polymerization is greater than the amount of fluid required for support of the fluidized bed and for adequate solids mixing in the fluidized bed. However, the fluid velocity in the reactor is limited due to the need to prevent excessive entrainment of solids in the

fluidizing gas stream as it exits the reactor.
Necessarily then, the amount of fluid which can be
circulated to remove heat is similarly limited.

One method of increasing the rate of heat
removal is to compress and cool the monomeric gases
recycled to the reactor to a point where a portion
thereof is condensed. The resulting liquid portion
is entrained in the recycled monomeric gas stream
and carried back to the reactor. Such operation
has been referred to as the "condensing mode" of
polymerization operation and is disclosed in prior
U. S. Application Serial Number 361,547, filed March
24, 1982 and in contemporaneously filed U.S. Patent
Application of J.M. Jenkins et al. (Attorney's
docket D-13,475-2), entitled "Improved Method for
Fluidized Bed Polymerization", both of which are
incorporated herein by reference. As there
disclosed, the employment of the condensing mode of
operation permits a reduction in the temperature of
the recycle stream, which, in combination with the
heat of vaporization of the liquid, results in a
marked increase in space-time- yield over that
obtainable in the "non-condensing mode" of operation
where the temperature of the recycled gas stream is
maintained above the dew point of the recycle
stream, the dew point being the temperature at which
liquid condensate begins to form in the gas stream.

Test results from a scale model bottom
reactor head and experience with a commercial
polymerization reactor have indicated that an open
nozzle-type reactor inlet is satisfactory for
successful operation of a fluidized bed reactor in
the condensing mode while a standpipe/conical

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cap-type reactor inlet is satisfactory for a
non-condensing mode of operation of the reactor.
The standpipe/conical cap-type inlet is not
satisfactory for a condensing mode of operation due
to liquid flooding or frothing in the bottom head, a
phenomenon experienced with commercial reactors at
relatively low levels of liquid in the recycle
stream. Conversely, the open nozzle-type inlet has
been found to be unsatisfactory for a non-condensing
mode of operation in a commercial reactor because of
excessive build-up of resin solids in the bottom
head, particularly around the inlet opening.

In the practical operation of large
commercial production reactors it is sometimes
desireable to switch from a non-condensing mode of
operation to a condensing mode of operation and vice
versa. To do this in the past, for the reasons
given above it has been necessary to shut down the
reactor to replace or alter the inlet to conform
with the requirements of the new mode of operation.
Reactor shutdown for the transition is undesirable
not only because of the maintenance costs associated
with the changeover, but because the downtime
results in significant production losses. For some
commercial reactors, the transition may be required
frequently, depending upon production schedules.
Consequently, it is highly desirable to have an
all-purpose reactor inlet configuration which
satisfies the requirements for both the condensing
and non-condensing modes of reactor operation.

The objectives of the invention, then, are
to (1) augment the production rates of fluidized bed

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reactors, (2) lower the cost for the maintenance
and/or operation of such reactors, and (3) provide
flexibility to enable such reactors to produce a
variety of polymers, e.g., polymers of ethylene and
heavier alpha olefins (copolymers and terpolymers)
and polymers of propylene (homopolymers and block or
random copolymers) at higher than conventional
production rates without incurring downtime
production losses. The flow deflector means of the
present invention assists in meeting these
objectives by providing a multi-purpose reactor
inlet configuration which eliminates the need for
reactor shutdown when converting from the condensing
mode to the non-condensing mode of operation and
vice versa.

In accordance with the present invention, a
fluidized bed polymerization reactor is provided
having distributor plate means below the fluidized
bed region defining a mixing chamber within the
reactor in the region below the distributor plate
means and one or more entry means for passage of
fluid into the reactor and through the mixing
chamber. At least one flow deflector means is
positioned below the distributor plate means and is
associated with at least one of the entry means.
The flow deflector means is adapted to provide at
least two fluid flow paths for fluid entering the
mixing chamber, a first fluid flow path along the
wall of the mixing chamber, and a second upwardly
oriented fluid flow path wherein, in operation,
solid particles, if any, in or coming into the

second flow path are carried upward; the wall of the
mixing chamber is swept to inhibit the build-up of
such solid particles; and the accumulation of liquid
in the mixing chamber is also inhibited.

The preferred flow deflector is an annular
flow deflector having aperture means for providing a
central, upwardly oriented fluid flow path and means
for providing peripheral flow around the flow
deflector means and along the wall of the mixing
chamber. By the operation of such a flow deflector
the fluid stream entering the mixing chamber through
the entry means is divided to form a flow path
through the aperture means and an outer peripheral
flow path between the deflector means and the wall
of the mixing chamber.

Figure 1 is a schematic illustration of a
preferred system for practicing the continuous
fluidized bed polymerization of monomers in
accordance with the invention;

Figure 2 is an elevational view in cross
section of the lower portion of the reactor
including the distributor plate means, the mixing
chamber and flow deflector means;

Figure 2A is a top view of the flow
deflector means as viewed along line 2a-2a of Figure
2;

Figure 3 is an enlarged partial elevational
view in cross section of the entry means and flow
deflector means of Figure 2;

Figure 4 is a top view of the distributor
plate as viewed along line 4-4 of Figure 2 showing
an orientation of cap means positioned thereon;

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Figure 5 is a partial elevational view in cross section of the entry portion of the reactor showing an alternative embodiment of the flow deflector means to that of Figure 3; and

Figure 6 is a partial elevational view in cross section of the entry portion of the reactor showing still another alternative embodiment of the flow deflector means.

In a continuous fluidized bed polymerization reactor a distributor plate is provided below the fluidized bed which supports the bed and provides gas distribution across the fluidized bed. The distributor plate also serves to define a mixing chamber in the region of the reactor below the distributor plate. The primary function of the mixing chamber is to ensure that the gases and entrained condensed liquid (if any) are distributed across the full cross-section of the chamber by the time the gas stream passes through the ports of the distributor plate at the base of the fluidized bed. Typically, when operating in a condensing mode of operation, some separation of the entrained liquid occurs on the surfaces of the fluid conduit through which the recycled stream enters into the reactor (primarily due to inertial impaction on the walls of the recycle conduit). To obtain sufficiently uniform entrainment and distribution it is important that any separated liquid present in the mixing chamber be re-entrained and mixed. This should be accomplished such that the gas stream passing through the ports of the

distributor plate means into the fluidized bed is of the desired uniformity. The flow deflector means of the present invention provides the desired distribution.

As shown schematically by arrows in Figures 2 and 3 of the drawings, in the preferred embodiment therein disclosed, the recycle gas stream is divided into a centrally disposed, upwardly oriented stream flowing through a central aperture in the annular flow deflector (annulus) and an outer stream passing peripherally or laterally around the annular flow deflector. The annular outer stream passes peripherally around the annular flow deflector and sweeps the wall of the mixing chamber; inhibiting the build-up of solid material (resin) thereon. It should be understood that a small amount of solids is generally entrained in the recycle fluid. As indicated by the arrows, the centrally disposed, upwardly oriented stream and the peripheral stream mix or commingle to provide the desired more uniform distribution of any liquid and/or solid material in the gas stream.

The parameters set out in Table 1 below serve to delineate operable conditions for providing the desired flow characteristics.

TABLE 1

Range	Preferred Value
(1) $0.1 \leq A_2/A_1 \leq 0.75$	0.3

wherein A_1 = the area through which the stream of

fluid passes in following the first flow path along the wall of the mixing chamber (curtain area), and

A_2 = the area of the aperture(s) through which the upwardly oriented stream flows in following the second flow path, and

wherein, for the preferred annular flow deflector means in Figures 2 and 3:

$$\text{area of aperture} = \frac{\pi d_1^2}{4}$$

where d_1 = the diameter of the central orifice; and

$$\text{curtain area} = \pi d_o h$$

where d_o is the outside diameter of the annular flow deflector and h is the minimum distance from the lower outer edge of the annular disc to the mixing chamber wall.

In defining the curtain area it is assumed that the supports or spacers 32a (see Figures 2 and 3) are of limited or negligible size in relation to the open area in the "curtain" area and can be ignored for purposes of the relationship set out above. That is, in the preferred mode disclosed in Figures 2 and 3, the spacers occupy a relatively small part of the curtain area.

Preferred
Value

$$2$$

$$(2) \quad 0.5 \leq \frac{Z}{\text{minimum clearance between the flow deflector and the inside wall of the mixing chamber}} \leq 5$$

wherein Z is the horizontal distance between the top inside edge of the entry pipe and the outer tip of the flow deflector, and wherein, for the annular flow deflector of Figures 2 and 3:

$$0.5 \leq \frac{d_o - d_e}{2h} \leq 5,$$

since, for the annular flow deflector, Z is $\frac{d_o - d_e}{2}$,

where d_o is as defined above and d_e is the diameter of the entry means (inlet pipe or recycle line 22 in Figure 2).

Preferred Range

$$(3) \quad H_v > 0.05\text{psi} \qquad H_v > 0.2\text{psi}$$

wherein H_v is the velocity head based on the total flow deflector flow area with the total flow deflector flow area defined as the sum of: the area of the first flow path along the wall of the mixing chamber (the curtain area) plus the area of the aperture(s) of the second flow path through which the upwardly oriented stream passes and wherein:

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$$H_v = \rho_g V_G^2/9266**,$$

where ρ_g is the density of the gas (lb/ft^3);
and the gas velocity (V_G) in ft/sec is:

$$V_G = 144 \ W_G/A_D \rho_g,$$

where W_G is the mass flow rate (lb/sec) of gas
entering the deflector and, for the system of
Figures 2 and 3, the total deflector open area,

$$A_D = \pi \ d_i^2/4 + \pi \ d_o h.$$

It is to be noted that d_i, d_o and h are
measured in inches.

Under such conditions of operation, the
resultant velocities of the central and annular or
peripheral streams and their relative mass flow
rates assure intimate mixing of the streams and
continued suspension of entrained liquid and solids
in the upwardly flowing gas stream entering into the
fluidized bed through the distribution plate means.
It has been found that there is no permanent
disentrainment of liquid droplets or solids out of
the gas streams under these operating conditions.

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** The cross-sectional area of the aperture(s)
(A_2) should not exceed about 2/3 of the cross-
sectional area of the entry means.

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Nor is there undesirable liquid flooding of the
mixing chamber or solids (resin) buildup therein,
which can result with disentrainment of liquid and
solids, respectively.

In accordance with the invention, by
maintaining the relationship between relative mass
flow rates and velocities in the upwardly oriented
fluid stream and the stream in the fluid flow path
along the wall, the desired positive flow in both
paths is obtained and the desired level of mixing in
the mixing chamber is achieved.

The mixing chamber should generally have a
length-to-width (diameter) ratio of up to about 1.5
and preferably from about 0.7 to about 1.0.
Similarly, the ratio of the diameter of the mixing
chamber to the diameter of the fluid entry means
(inlet or conduit) to the reactor is generally less
than or equal to about 10:1, preferably in the range
of from about 5:1 to about 8:1.

Additionally, when operating in the
condensing mode of operation, the ratio of the
superficial gas velocity (U_m), in the mixing
chamber, to the terminal gas velocity in the
condensing mode of operation in the lower mixing
chamber (U_1), as defined below, should preferably
be at least 0.18:1.

A ratio of U_m/U_1 of at least 0.18:1
reduces the likelihood of flooding and frothing in
the lower regions of the mixing chamber. U_m
represents the superficial gas velocity in the
chamber and U_1 represents the limiting gas
velocity, i.e., the gas velocity above which drops
of entrained liquid are subdivided and the entrained

liquid is carried upward with the gas stream. The latter velocity may be represented (in dimensionally consistent units) as follows:

$$U_1 = 2.0(g\sigma_1\Delta\rho/\rho_g^2)^{0.25}$$

where: g is the gravitational acceleration;

σ_1 is the surface tension of the liquid;

$\Delta\rho$ is the density difference between the liquid and the gas; and

ρ_g is the density of the gas.

In general for the condensing mode of operation it has been determined that the weight fraction of condensed liquid entrained in the recycle gas stream may range up to about 0.2 (20 weight percent), preferably from about 2 to about 20 weight percent, the particular weight percent being dependent on the particular polymer being produced.

A fluidized bed reaction system which is particularly suited for the production of polyolefin resins in accordance with the present invention is illustrated in the drawings. With specific reference to Figure 1, reactor 10 comprises a reaction zone 12 and a velocity reduction zone 14.

In general, the height-to-diameter ratio of the reaction zone will lie in the range of from

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about 2.7:1 to about 4.6:1. The range can vary to
larger or smaller ratios dependent upon the desired
production capacity. The cross-sectional area of
the velocity reduction zone 14 is typically within
the range of about 2.6 to about 2.8 multiplied by
the cross-sectional area of the reaction zone 12.

The reaction zone 12 includes a bed of
growing polymer particles, formed polymer particles
and a minor amount of partially or totally activated
precursor composition and/or catalyst (hereinafter
collectively referred to as catalyst), all fluidized
by the continuous flow of polymerizable and
modifying gaseous components in the form of make-up
feed and recycle fluid through the reaction zone.
To maintain a viable fluidized bed, the superficial
gas velocity (SGV) through the bed must exceed the
minimum flow required for fluidization, which is
typically from about 0.2 to about 0.5 feet per
second. Preferably, the SGV is at least 0.2 feet
per second above the minimum flow required for
fluidization, i.e., typically from about 0.4 to
about 0.7 feet per second. Ordinarily, the SGV will
not exceed 5.0 feet per second and it is usually no
more than 2.5 feet per second.

Particles in the bed help to prevent the
formation of localized "hot spots" and to entrap and
distribute the particulate catalyst throughout the
reaction zone. Accordingly, on start up, the
reactor is charged with a base of particulate
polymer particles before gas flow is initiated.
Such particles may be the same as the polymer to be
formed or different. When different, they are
withdrawn with the desired newly formed polymer

particles as the first product. Eventually a
fluidized bed consisting of desired polymer
particles supplants the start-up bed.

The catalysts used are often sensitive to
oxygen, thus the catalyst used to produce polymer in
the fluidized bed is preferably stored in a
reservoir 16 under a blanket of a gas which is inert
to the stored material, such as nitrogen or argon.

Fluidization is achieved by a high rate of
fluid recycle to and through the bed, typically on
the order of about 50 times the rate of feed of
make-up fluid. The fluidized bed has the general
appearance of a dense mass of individually moving
particles as created by the percolation of gas
through the bed. The pressure drop through the bed
is equal to or slightly greater than the weight of
the bed divided by the cross-sectional area. It is
thus dependent on the geometry of the reactor.

Make-up fluid is fed to the reactor system
at point 18 via recycle line 22. The composition of
the recycle stream is measured by a gas analyzer 21
and the composition and amount of the make-up stream
is then adjusted accordingly to maintain an
essentially steady state gaseous composition within
the reaction zone.

The gas analyzer is a conventional gas
analyzer which operates in conventional manner to
indicate recycle stream composition and which is
adapted to regulate the feed and is commercially
available from a wide variety of sources. The gas
analyzer 21 can be positioned to receive gas from a
point between the velocity reduction zone 14 and the

heat exchanger 24, preferably, between the
compressor 30 and heat exchanger 24.

If desired, other additives may be added to
the recycle line 22 through line 40 from a suitable
dispenser 38.

To ensure complete fluidization, the
recycle stream and, where desired, part of the
make-up stream are returned through recycle line 22
to the reactor at point 26 below the bed.
Preferably there is a gas distributor plate 28 above
the point of return to aid in fluidizing the bed
uniformly and to support the solids particles prior
to startup or when the system is shut down. The
stream flowing upwardly through the bed absorbs the
heat of reaction generated by the polymerization
reaction.

The portion of the gaseous stream flowing
through the fluidized bed which did not react in the
bed becomes the recycle stream which leaves the
reaction zone 12 and passes into a velocity
reduction zone 14 above the bed where a major
portion of the entrained particles drop back into
the bed thereby reducing solid particle carryover.

In very general terms, a conventional
fluidized bed process for producing resins,
particularly polymers produced from monomers, is
practiced by passing a gaseous stream containing one
or more monomers continuously through a fluidized
bed reactor under reactive conditions and in the
presence of catalyst at a velocity sufficient to
maintain the bed of solids particles in a suspended
condition. The gaseous stream containing unreacted
gaseous monomer is withdrawn from the reactor

continuously, compressed, cooled and recycled into the reactor. Product is withdrawn from the reactor and make-up monomer is added to the recycle stream. In the course of sweeping the gaseous stream through the fluidized bed reactor to maintain the bed in a suspended condition, a minor portion of the solids particles present in the bed may be carried out of the reactor with the gaseous stream which is recycled to the reactor. Since these particles are hot and contain catalyst, they will continue to grow by further reaction with monomer gas as they are carried through the recycle system, potentially causing problems by settling out and agglomerating into a solid mass or sticking to the walls of the recycle line and heat exchanger. This may ultimately lead to plugging of the recycle lines or heat exchanger, necessitating shutdown. Accordingly, it is important to minimize the carryover of particles in the recycle stream.

While virtually all solid particle carryover can be eliminated, the penalty to accomplish this is substantially increased capital cost for auxiliary equipment, e.g., cyclones, and the increased cost for the maintenance and operation of this auxiliary equipment. Since a minor amount of solid particle carryover in the recycle stream is manageable, it is preferred to accept a minimal amount of solid particle carryover rather than to eliminate it completely. However, when operating in a condensing mode, in accordance with the subject invention, the additional problem of "mud", as is discussed in detail below, can occur.

The polymer-forming reaction is exothermic,
making it necessary to maintain in some fashion the
temperature of the gas stream inside the reactor at
a temperature not only below the resin and catalyst
degradation temperatures but at a temperature below
the fusion or sticking temperature of resin
particles produced during the polymerization
reaction. This is necessary to prevent plugging of
the reactor due to rapid growth of polymer chunks
which cannot be removed in a continuous fashion as
product. It will be understood, therefore, that the
amount of polymer that can be produced in a
fluidized bed reactor of a given size in a specified
time period is directly related to the amount of
heat which can be withdrawn from the fluidized bed.

When operating in the condensing mode of
operation, the recycle gas stream is intentionally
cooled to a temperature below the dew point of the
recycle gas stream to form a mixture comprising a
liquid phase and a gas phase and which may also
contain a minor amount of solid particles.

When operating in the condensing mode, it
may be desirable in some instances to raise the dew
point of the recycle gas stream to further increase
heat removal. The recycle stream dew point can be
increased by: (1) raising the operating pressure of
the reaction system, (2) increasing the
concentration of condensible fluids in the recycle
stream, and/or (3) reducing the concentration of
non-condensible gases in the recycle stream. For
example, the dew point of the recycle stream may be
increased by the addition of a condensible fluid to
the recycle stream which is inert to the catalyst,

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reactants, and the products of the polymerization
reaction. The fluid can be introduced into the
recycle stream with the make-up fluid or by any
other means or at any other point in the system.
Examples of such fluids are saturated hydrocarbons,
such as butanes, pentanes or hexanes.

A primary limitation on the extent to which
the recycle gas stream can be cooled below the dew
point is in the requirement that the gas-to-liquid
ratio be maintained at a level sufficient to keep
the liquid phase of the mixture in an entrained or
suspended condition until the liquid is vaporized.
It is also necessary that the velocity of the
upwardly flowing fluid stream immediately above the
gas distributor plate be sufficient to maintain the
fluidized bed in a suspended condition.

The liquid content of the recycle stream
can be quite high, but as a general rule the
quantity of condensed liquid contained in the gas
phase should not exceed about 20 weight percent at
the point of passage through the distributor plate
(based on the total weight of the recycle stream).
When the liquid content is below 2 weight percent
the benefit obtained is limited.

To the extent that solid particles are
carried over in the gaseous stream exiting the
reactor, it is important that the amount of liquid
present in the recycle stream, when operating in the
condensing mode, be sufficient to avoid forming
"mud". Undesirable "mud" may result from wetting of
solid particles, agglomeration, and disentrainment
resulting in accumulation and deposition in regions
of relatively low velocity in the system, e.g., in

the heat exchanger or elsewhere in the recycle
lines. The amount of solids in the gaseous stream
exiting the reactor is typically small, e.g., from
about 0.1 to about 0.5 weight percent (based on the
total weight of the stream). However, larger
amounts, on the order of 1 weight percent or higher,
may occur. Since the ratio of liquid to solid
particles at which mud can form is somewhat variable
(based, it is believed, at least in large part on
particle shape and distribution), the weight ratio
of liquid to solid particles in the recycle stream
is maintained at not less than about two to one,
preferably not less than about five to one, and more
preferably not less than 10 to 1 to avoid this
potential problem. The latter higher ratios provide
protection against aberrations in operation which
may result in transitory higher solids carryover in
the gaseous stream exiting the reactor.

An excess of liquid serves to prevent the
accumulation of solids in the system at any point in
the system where they might otherwise settle out and
to keep the system washed clean. In no event should
the ratio fall below about two to one when the
amount of liquid in the entering recycle stream is
in the desired operating range of from 2 to about 20
weight percent. When operating in the
non-condensing mode, i.e., without liquid in the
recycle or with very low levels of liquid present,
the ratio of liquid to solids particles in the
recycle stream is not significant since the solids
will not be wetted to any significant extent and the
formation of mud is not a problem.

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The recycle stream is compressed in a compressor 30 and then passed through a heat exchange zone where the heat of reaction is removed from the recycle stream before it is returned to the bed. The heat exchange zone may be a conventional heat exchanger 24 which can be of the horizontal or vertical type. The recycle stream exiting the heat exchange zone is returned to the reactor at its base 26 and to the fluidized bed through mixing chamber 26a and gas distributor plate 28. In the preferred embodiment illustrated in Figures 1-3, the annular disc deflector means is positioned at a stand-off distance from the inlet to the reactor at the base of the mixing chamber of 26a.

The preferred annular flow deflector means, as shown in Figures 2 and 3 of the drawings, comprises an annulus 32 supported at a standoff distance (h) above the reactor inlet 26 by spacers 32a divides the entering recycle stream into a central upward flow stream 33 and a peripheral annular flow stream 33a along the lower side wall of the reactor. The streams mix and pass thic__h the protective screen 27, the holes or ports 29 of the distributor plate 28, and the angle caps 36a and 36b, secured to the upper surface of the distributor plate, and then into the fluidized bed. The flow is a mixture of gas and generally a small amount of solid particles (resin) for a non-condensing mode of reactor operation. For a condensing mode of reactor operation, the flow is a mixture of gas, liquid droplets, and generally some solid particles (resin).

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The central upper flow stream 33 in the
mixing chamber 26a assists in the entrainment of
liquid droplets in the bottom head or mixing chamber
and in carrying the entrained liquid to the
fluidized bed section during a condensing mode of
reactor operation. The peripheral flow 33a assists
in limiting the build-up of solid particles in the
bottom head during both modes of reactor operation
because it sweeps the inner surface of the reactor
wall. The peripheral flow also contributes to the
re-atomization and re-entrainment of any liquid
which may be disentrained at the wall or accumulate
at the bottom of the mixing chamber during a
condensing mode operation, especially with higher
levels of liquid in the system. By providing both
central upward and outer peripheral flow in the
mixing chamber, the flow deflector means 32 permits
a reactor to be operated in either the condensing or
the non-condensing mode without the problems of
liquid flooding or excessive build-up of resin at
the bottom of the reactor.

The temperature of the bed is basically
'·pendent on three factors: (1) the rate of catalyst
injection which controls the rate of polymerization
and the attendant rate of heat generation, (2) the
temperature of the gas recycle stream, and (3) the
volume of the recycle stream passing through the
fluidized bed. Of course, the amount of liquid
introduced into the bed either with the recycle
stream and/or by separate injection also affects the
temperature since this liquid vaporizes in the bed
and serves to reduce the temperature. Normally the

rate of catalyst injection is used to control the
rate of polymer production. The temperature of the
bed is controlled at an essentially constant
temperature under steady state conditions by
constantly removing the heat of reaction. No
noticeable temperature gradient appears to exist
within the upper portion of the bed. A temperature
gradient will exist in the bottom of the bed in a
layer or region extending above the distribution
plate, e.g., for about 6 to 12 inches, as a result
of the difference between the temperature of the
inlet fluid and the temperature of the remainder of
the bed. However, in the upper portion or region
above this bottom layer the temperature of the bed
is essentially constant at the maximum desired
temperature.

Good gas distribution plays an important
role in the efficient operation of the reactor. The
fluidized bed contains growing and formed
particulate polymer particles, as well as catalyst
particles. As the polymer particles are hot and
possibly active, they must be prevented from
settling for, if a quiescent mass is allowed to
exist, any active catalyst present will continue to
react and can cause fusion of the polymer particles
resulting, in an extreme case, in the formation of a
solid mass in the reactor which can only be removed
with great difficulty and at the expense of an
extended downtime. Since the fluidized bed in a
typical commercial size reactor may contain many
thousand pounds of solids at any given time, the
removal of a solid mass of this size requires a

substantial effort. Diffusing recycle fluid through
the bed at a rate sufficient to maintain
fluidization through the bed is, therefore,
essential.

Gas distributor plate 28 is a preferred
means for achieving good gas distribution. It may
be a screen, slotted plate, perforated plate, a
plate of the bubble-cap type, or the like. The
elements of the plate may all be stationary or the
plate may by of the mobile type disclosed in U. S.
Patent 3,298,792. Whatever its design, it must
diffuse the recycle fluid through the particles at
the base of the bed to keep the bed in a fluidized
condition and also serve to support a quiescent bed
of resin particles when the reactor is not in
operation. Preferably, a protective screen 27 is
positioned below the distributor plate 28 to reduce
the likelihood of the plate being plugged by resin
chips impinging on the plate as the gas recycle
stream carries the chips upward.

The preferred type gas distributor plate
28 is generally of the type which is fabricated from
metal and wiuch has holes distributed across its
surface. The holes are normally of a diameter of
about one-half inch. The holes extend through the
plate and over the holes there are positioned angle
irons identified by reference numerals 36a and 36b
in Figure 1 which are fixedly mounted to plate 28.
Alternate rows of angle irons are oriented at angles
to each other, preferably at 60 degrees, in
alternate parallel alignment as shown in Figure 4.
They serve to distribute the flow of fluid along the

surface of the plate so as to avoid stagnant zones
of solids. In addition, they prevent resin
particles from falling through the holes when the
bed is settled or quiescent.

The fluid bed reactor may be operated at
pressures of up to about 1,000 psig and, for poly-
olefin resin production, preferably is operated at a
pressure of from about 250 psig to about 500 psig.

The partially or totally activated catalyst
is injected intermittently or continuously into the
bed at a desired rate at a point 42 which is above
the distributor plate 28. Preferably, the catalyst
is injected at a point in the bed where good mixing
with polymer particles occurs.

The catalyst can be injected into the
reactor by various techniques. For ethylene
polymerization it is preferred to continuously feed
the catalyst into the reactor utilizing a catalyst
feeder as disclosed, e.g., in U.S. Patent
3,779,712. The catalyst is preferably fed into the
reactor at a point 20 to 40 percent of the reactor
diameter away from the reactor wall and at a height
above the bottom of the bed equal to from about 5 to
about 30 percent of the height of the fluidized bed.

A gas which is inert to the catalyst, such
as nitrogen or argon, is preferably used to carry
the catalyst into the bed.

The rate of polymer production in the bed
depends on the rate of catalyst injection and the
concentration of monomer(s) in the recycle stream.
The production rate is conveniently controlled by
simply adjusting the rate of catalyst injection.

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As noted above, in order to minimize the
problem of distributor plate pluggage, a mesh screen
27 is preferably installed below the plate to
prevent the drilled holes in the plate from being
plugged by resin chips (agglomerated solids) from
the recycle stream.

On discharge of particulate polymer product
from the reactor 10, it is desirable, and
preferable, to separate fluid from the product and
to return the fluid to the recycle line 22. There
are numerous ways known to accomplish this. One
system is shown in the drawings. Thus, fluid and
product leave reactor 10 through line 44 and enter
the product discharge tank 46 through a valve 48
which is designed to have minimum restriction to
flow when opened, e.g., a ball valve. Positioned
above and below product discharge tank 46 are
conventional valves 50, 52 with the latter being
adapted to provide passage of product into a product
surge tank 54. Product surge tank 54 has venting
means illustrated by line 56 and gas entry means
illustrated by line 58. Also positioned at the base
of product surge tank 54 is a discharge valve 60
which, when in the open position, discharges product
for conveying to storage. Valve 50, when in the
open position, releases fluid to surge tank 62.
Fluid from the product discharge tank 46 is directed
through a filter 64 and thence through a surge tank
62, a compressor 66 and into recycle line 22 through
line 68.

In a typical mode of operation, valve 48 is
open and valves 50, 52 are in a closed position
Product and fluid enter product discharge tank 46.

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Valve 48 closes and the product is allowed to settle
in product discharge tank 46. Valve 50 is then
opened, permitting fluid to flow from product
discharge tank 46 to surge tank 62 from which it is
continually compressed back into recycle line 22.
Valve 50 is then closed and valve 52 is opened and
product in the product discharge tank 46 flows into
the product surge tank 54. Valve 52 is then
closed. The product is purged with inert gas,
preferably nitrogen, which enters the product surge
tank 54 through line 58 and is vented through line
56. Product is then discharged from product surge
tank 54 through valve 60 and conveyed through line
20 to storage.

Another more preferred product discharge
system which may be employed in the alternative is
that disclosed and claimed in the copending U.S.
Patent Application of Robert G. Aronson, filed
July 28, 1981, Serial No. 287,815 and entitled
"Fluidized Bed Discharge Process" (published as EPA
Application No. 0071430 on February 9, 1983). Such
a system employs at least one pair of tanks (and
more preferably two pairs of tanks in parallel)
comprising a settling tank and a transfer tank
arranged in series and having the separated gas
phase returned from the top of the settling tank to
a point in the reactor near the top of the fluidized
bed. This alternative preferred product discharge
system obviates the need for a recompression line
64, 66, 68, as shown in the system of the drawing.

In the basic embodiment of the system for
conducting the process of the invention, the
reaction vessel contains a bed of fluidized solid

particles with a gas stream entering an inlet line
at the bottom and leaving through a vent line at the
top. A vented settling tank, located externally and
preferably below the fluidized bed, is connected to
the bed by a discharge line and a vent line. The
vent line is directly connected to the reaction
vessel near the top level of fluidized bed and the
solids discharge line is connected to the lower part
of the vessel, preferably near the distributor
plate. A transfer tank is located below and is
connected to the bottom of the settling tank by a
line and is connected to downstream processing
equipment through a discharge line. Initially, the
reaction vessel and settling and transfer tanks are
isolated from each other and from downstream
equipment by valves. Solids and gas are discharged
from the reaction vessel into the settling tank by
opening discharge and vent line valves while keeping
the settling tank outlet valve closed. The settling
tank pressure initially increases to approximately
that in the bottom of the reaction vessel and then
the differential pressure across the fluidized bed
of solids is the driving force ror solids and gas to
flow through the discharge line. A portion of this
fluidizing gas and solids flows through discharge
line to the settling tank because of the lower
resistance flow path compared to the flow path
through the fluidized bed of solids. The solids and
gas separate in the settling tank and the gas is
returned to the reaction vessel through the vent
line by displacement with additional solids and gas
entering the settling tank. After the settling tank
fills with settled solids and some gas, the settling

tank is isolated from the reaction vessel by closing
the discharge and vent line valves. The solids are
then transferred, by pressure differential and by
gravity, from the settling tank to the transfer tank
through a line by opening the valve therein. With
the solids in the transfer tank and the tank
pressures equalized, the line valve is closed. The
settling tank is now ready to start another
discharge cycle or it can wait until the transfer
tank is finished transferring the solids to
downstream equipment. Solids are then transported
from the transfer tank to downstream equipment at
lower pressure by opening the outlet valve. Solids
transfer can be by conventional solids handling
equipment or by high pressure conveying using the
pressurized gas that is entrained with the solids
(additional gas may be required). After the solids
are transported from the transfer tank, the outlet
valve is closed and the transfer tank is ready for
another cycle.

In an alternate and preferred embodiment,
solids are intermittently discharged from a high
pressure vessel containing a fluidized bed of solids
using a process that employs two pairs of the basic
settling and transfer tanks operated in parallel and
operating sequentially with gas vented from the
solids prior to discharging the solids to a lower
pressure. A first vented settling tank serves to
receive solids and gas discharged from the fluidized
bed. After the tank fills with solids, some of the
gas entrained with the solids is vented to a second
settling tank (in the parallel pair system) which

serves as a temporary gas receiver, later indirectly venting to the reaction vessel. The solids are then transferred from the settling tank to a lower pressure transfer tank while minimizing gas losses. The discharge operation is continued alternating between parallel settling tanks in receiving a solids and gas stream from the fluidized bed.

The fluidized-bed reactor is equipped with an adequate venting system (not shown) to allow venting the bed during startup and shutdown. The reactor does not require the use of stirring and/or wall scraping. The recycle line 22 and the elements therein (compressor 30, heat exchanger 24) should be smooth surfaced and devoid of unnecessary obstructions so as not to impede the flow of recycle fluid or entrained particles.

EXAMPLES

A commercial fluidized bed olefin polymerization reactor, in the bottom of which an annular flow deflector means of the type shown in Figures 2 and 3 of the drawings was employed, has been operated in both condensing and non-condensing modes without problems. The dimensions were as follows: d_m = 11.5 feet (3.51 m) mixing chamber diameter; L = 8.3 feet (2.54 m) mixing chamber height; d_e = 23 inches (0.58 m) inlet diameter; d_o = 38 inches (0.97 m) outer deflector diameter; d_i = 13.9 inches (0.35 m) inner deflector diameter; and h = 3.9 inches (0.10 m) deflector stand off distance. Four spacers were used to support the annulus and to maintain the stand off distance clearance (h) between the outer edge of the annulus and the reactor bottom. For this annular

disc, A_2/A_1 was 0.33, Z/h was 1.9 and H_v was 1.0 psi.

The reactor has been used to produce ethylene copolymers in both the condensing and the non-condensing modes and ethylene homopolymers in the non-condensing mode. Depending upon the products produced, the reactor conditions can be, for example:

Reactor temperature, T_{bed}:89-95 degrees C;

Reactor pressure, P_{bed}:300-305 psig (20.7-21.0 bar);

U_s: 1.8-2.3 feet per second (0.55-0.70 meters per second) where Us is the superficial gas velocity in the fluidized bed

Fluidized bed height, H_{fb}:39 feet (11.9 meters);

Maximum rate of condensation experienced at the reactor inlet, W_{max}:11 weight percent;

Reactor production rate: 21,000-40,000 pounds per hour (9,526-18,145 kilograms per hour).

No problem of reactor operation and no adverse effect on the quality of products, attributable to the flow deflector means were observed. Even at the highest rate of condensation (11 weight percent at the reactor inlet) that the reactor experienced, there was no reactor operational instability during the run. This indicated that, at that level of condensation, the liquid was well entrained and carried into the fluidized bed as droplets without accumulating in the bottom head and flooding. The reactor was

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inspected often and no fouling due to excessive
build-up of resin solids was observed. The inner
surfaces were found to be clean, even cleaner than
in the case of utilizing the conventional standpipe/
conical cap-type reactor inlet. Accordingly, the
use of flow deflector means in accordance with this
invention provides a method for operating in both
the condensing and non-condensing modes of operation
without adversely affecting the properties or
quality of the products.

The following Tables set forth examples of
the production by polymerization of ethylene
polymers employing either the condensing or
non-condensing mode of operation. The commercial
polymerization reactor employed was the reactor
described immediately above. The runs were carried
out to produce the products listed in Table 2 in the
mode of operation therein stated and with the amount
of condensation stated. The full operating
conditions for the two runs are in the data set
forth in Table 3.

TABLE 2

Run	Product	Mode of Operation	Amount of Liquid in the gas stream at the Reactor Inlet (Weight Percent)
I	Hexene copolymer of ethylene	Condensing	9-10
II	Butene copolymer of ethylene	Non-condensing	0

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TABLE 3

Run	I	II
ΔP bottom* (psi)	5.2	5.0
ΔP cooler (psi)	6.0	6.5
T_{bed} (°C)	95.0	89.0
P_{bed} (psig)	305.0	305.0
U_i (ft/sec) (inlet velocity)	79.2	79.2
U_s (ft/sec) (superficial gas velocity in the fluidized bed)	2.2	2.2
L_{fb} (ft) (fluidized bed height)	39.0	39.0
L (ft) (diffusion mixing chamber height)	8.33	8.33
D_m (ft) (diffusion mixing chamber diameter)	11.5	11.5
Production rate (KPPH)	40.0	22.0
Space time yield (lb/ft^{3}hr)	9-10.0	5.4
MI (melt index)	0.83	100.0
MFR (melt flow ratio)	25.0	27.0
ρ(g/cc)	0.926	0.931
Ash (%)	0.042	0.040

TABLE 3 (Continued)

Run	I	II
ρ_b (lbm/ft^3) (bulk density)	27.0	24.5
APS (in) (average particle size)	0.027	0.0226

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*ΔP bottom denotes pressure drop across the flow deflector means plus the pressure drop across the distributor plate.

Note: I   = Hexene copolymer of ethylene
         II = Butene copolymer of ethylene

- - - - - - - - -

As shown in Figure 5 and 6 of the drawings, the flow deflector means need not be planar and oriented in a horizontal plane (as is the preferred flow deflector means illustrated in Figures 2 and 3). For example, the flow deflector means illustrated in Figures 5 and 6 may be substituted for the preferred flow deflector means in Figure 2 in which event the flow deflector means are convex and concave with respect to the distributor plate, respectively. In a manner similar to that of the flow deflector means illustrated in Figures 2 and 3, the flow deflector means of Figures 5 and 6 when positioned over the bottom entry means 26 both provide a first fluid flow path along the wall of the mixing chamber and a second central, upwardly oriented fluid flow path through the central aperture of the flow deflector means.

The vertical height or thickness of the deflector means is not critical and need be only as thick as structural demands of reactor construction require. A relatively thin deflector means will function with essentially the same result as one of considerable thickness. Thus, in addition to the lack of criticality in the generally horizontal arrangement of the flow deflector means and its ability to operate successfully both with or without a concave or convex shape, the thickness of the flow deflector means has also not been found to be critical.

Although the base or bottom head of a commercial fluidized bed reactor is generally either an ellipsoidal or hemispherical dish which is connected with a straight section, other configurations may also be used with this invention. For example, the bottom head of the reactor may have a diverging conical shape and still be able to use the flow deflector concept. It is to be noted that the flat and convex annular flow deflectors as illustrated in the drawings are preferable to the concave annular flow deflectors for the ellipsoidal and hemispherical bottom reactors.

The subject invention finds use in a variety of fluidized bed polymerization reactor systems. It has applicability in the operation of fluidized bed reactors where it is necessary or desirable to switch from a condensing mode to a non-condensing mode of operation or vice versa. The invention finds particular applicability in the fluidized bed polymerization of polyolefins, e.g., polyethylene and polypropylene and their comonomers.
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CLAIMS:

1.   A fluidized bed polymerization
reactor having distributor plate means below the
fluidized bed region defining a mixing chamber
within said reactor in the region below said
distributor plate means and one or more entry means
for passage of fluid into said reactor and through
said mixing chamber, characterized in that it comprises
at least one flow deflector means positioned below
said distributor plate means and associated with at
least one of said entry means, said flow deflector
means adapted to provide at least two fluid flow
paths for said fluid entering said mixing chamber, a
first fluid flow path along the wall of said mixing
chamber, and a second upwardly oriented fluid flow
path wherein, in operation, solid particles, if any,
in or coming into said second flow path are carried
upward; the wall of said mixing chamber is swept to
inhibit the build-up of such solid particles; and
the accumulation of liquid in said mixing chamber is
also inhibited.

2.   A fluidized bed polymerization
reactor having distributor plate means below the
fluidized bed region defining a mixing chamber
within said reactor in the region below said
distributor plate means and one or more entry means
at or near the base of said reactor for passage of
fluid into said reactor and through said mixing
chamber, characterized in that it comprises at least
one flow deflector means positioned in said mixing
chamber and associated with at least one of said
entry means, said flow deflector means adapted to

provide at least two fluid flow paths for said fluid
entering said mixing chamber, a first fluid flow
path along the wall of said mixing chamber, and a
second upwardly oriented fluid flow path wherein, in
operation, solid particles, if any, in or coming
into said second flow path are carried upward; the
wall of said mixing chamber is swept to inhibit the
build-up of such solid particles; and the
accumulation of liquid in said mixing chamber is
also inhibited.

3. A fluidized bed reactor in accordance
with claim 2 wherein said flow deflector means is
generally annular in shape and positioned over said
entry means at a stand-off distance whereby said
first fluid flow path passes through the curtain
area below said flow deflector means and said second
flow path passes through aperture means in said flow
deflector means.

4. A fluidized bed reactor in accordance
with claim 3 wherein said flow deflector means is
positioned generally perpendicular to the axis of
said reactor.

5. A fluidized bed reactor in accordance
with claim 3 wherein said flow deflector means is
positioned concave relative to the said distributor
plate means.

6. A fluidized bed reactor in accordance
with claim 3 wherein said flow deflector means is
positioned convex relative to the said distributor
plate means.

7. A fluidized bed reactor in accordance with any one of claims 3 to 6, wherein said mixing chamber has a length-to-diameter ratio of up to about 1.5, preferably between about 0.7 and about 1.0.

8. A fluidized bed reactor in accordance with any one of claims 3 to 7, wherein:

the ratio of the area of said aperture means (A_2) to the area of said curtain area (A_1) satisfies the relationship:

$$0.1 \leq A_2/A_1 \leq 0.75; \text{ and}$$

$$0.5 \leq (d_o - d_e)/2h \leq 5$$

where d_o is the diameter of said annular flow deflector, d_e is the diameter of said entry means, and h is the minimum distance from the lower outer edge of said annular flow deflector to the mixing chamber wall and wherein, in operation, $H_v >$ 0.05 psi, where H_v is the velocity head based on total flow deflector flow area.

9. A fluidized bed reactor in accordance with claim 8, wherein the ratio A_2/A_1 is about 0.3 and $(d_o - d_e)/2h$ is about 2 and wherein, in operation, $H_v > 0.2$ psi.

10. A fluidized bed reactor in accordance with any one of claims 2 to 9, wherein said annular flow deflector means has dimensions such that:

$H_v > 0.05$ psig, preferably > 0.2 psig, wherein H_v is the velocity head based on total flow deflector flow area.

11. A fluidized bed polymerization reactor which comprises:

    A. a generally cylindrical vessel:

    B. distributor plate means positioned in said vessel, generally perpendicular to the axis of said vessel, defining a fluidized bed region above said distributor plate means and a mixing chamber region below said distributor plate means;

    C. inlet means for continuously introducing a stream of polymerizable fluids into said mixing chamber at a gas velocity sufficient to maintain particles in said fluidized bed region in suspended and fluidized condition;

    D. outlet means for continuously removing unreacted polymerizable gases from said fluidized bed region;

    E. catalyst injection means for introducing polymerization catalyst into said fluidized bed region;

    F. product removal means for removing solid particulate polymer product from said fluidized bed region; and

    G. flow deflector means associated with said inlet means, adapted to divide said stream of polymerizable fluid flowing through said inlet means into two or more streams flowing in at least two fluid flow paths of which at least one flow path is directed upwardly and outwardly along the wall of said mixing chamber and at least one flow path is directed generally upwardly along a central axis of said mixing chamber, the division of said stream being such that the velocity of fluid flowing upwardly along the walls of said mixing chamber is sufficient to (1) entrain solid particles and maintain particles in

the fluid entrained and (2) inhibit the
build-up of solid particulate polymer
product on the wall of said mixing chamber
and the velocity of fluid flowing upwardly
along a central axis of said mixing chamber
being sufficient to prevent solid
particulate polymer product from falling
into said inlet means; the accumulation of
liquid in the said mixing chamber is. also
inhibited; and the combined overall
velocity and direction of said fluid
flowing in said flow paths being such that
sufficient mixing is provided to ensure the
general uniformity and distribution of
polymerizable fluids entering the said
fluidized bed region.

12.   A method of operating a fluidized bed
polymerization reactor having distributor plate means
positioned below said bed to provide both distribution
of fluid and physical support for said bed;   entry
means at or near the base of said reactor for passage
of fluid into said reactor and a mixing chamber po-
sitioned between said distributor plate means and said
entry means and wherein a fluid stream comprising a
gas is continuously introduced into said reactor,
passed upward through said mixing chamber, through
said distributor plate means and through said flui-
dized bed, characterized in that it comprises divi-
ding the entering fluid stream into at least two
streams, a first stream flowing in an upward and out-
ward peripheral flow path along the wall of said mi-
xing chamber, and a second stream flowing upward in
a fluid flow path generally perpendicular to said
distributor plate means.

13.   A method of operating a fluidized bed
polymerization reactor having distributor plate

means positioned below said bed to provide both
distribution of fluid and physical support for said
bed; entry means at or near the base of said reactor
for passage of fluid into said reactor; generally
annular fluid flow deflector means associated with
said entry means and a mixing chamber positioned
between said distributor plate means and said entry
means and wherein a fluid stream comprising a gas is
continuously introduced into said reactor through
said entry means, passed upward through said mixing
chamber, through said distributor plate means and
through said fluidized bed, characterized in that it
comprises dividing the entering fluid stream into at
least two streams, a first stream flowing upward and
outward in a fluid flow path peripherally between
said annular deflector means and the wall of said
mixing chamber and a second stream flowing in a
central upward fluid flow path through said annular
flow deflector means, the combined velocity and
orientation of all fluid flow paths being capable of
entraining and passing solid particles and liquid
out of said mixing chamber through said distributor
plate means and into said fluidized bed.

14. A method of operating a fluidized bed
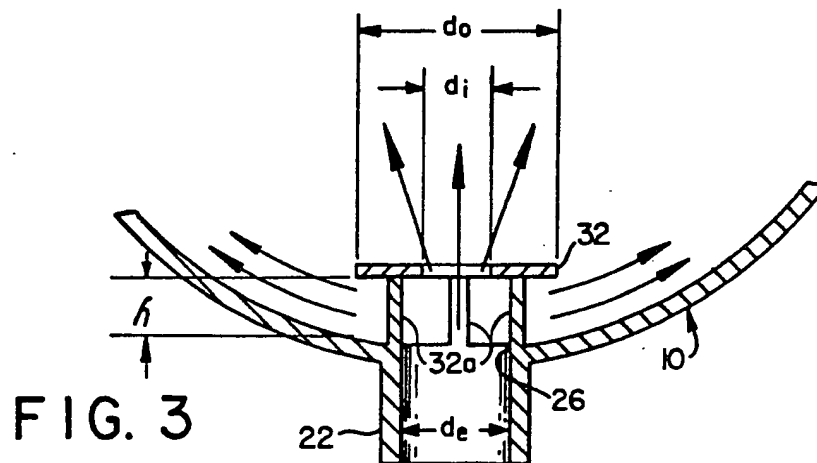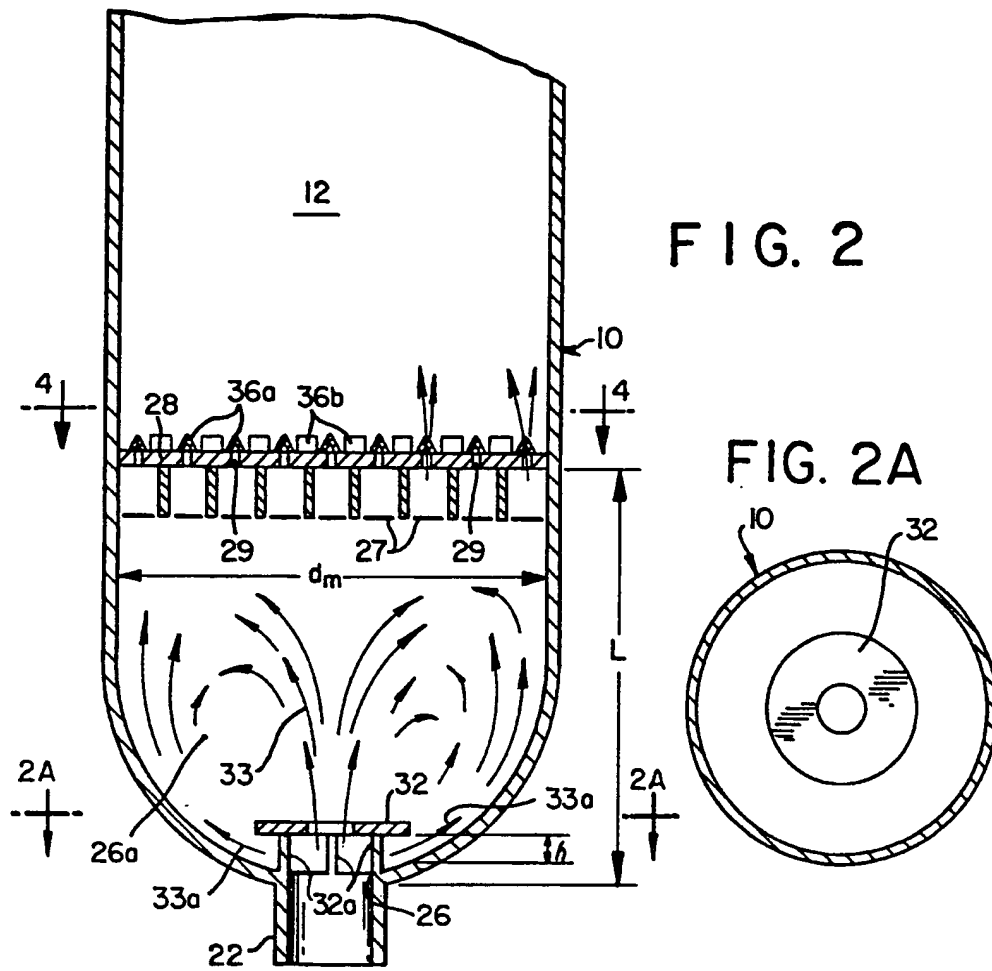polymerization process which comprises:

A. continuously introducing a
polymerizable fluid stream through an inlet
into a mixing chamber at a velocity
sufficient to maintain a particle bed in a
fluidized and suspended condition;

B. dividing said fluid stream into at
least two streams flowing in at least two
fluid flow paths; directing the fluid in at
least one of said flow paths upwardly and
outwardly along the wall of said mixing
chamber and directing the fluid in at least
one of said flow paths upwardly along a
central axis of said mixing chamber,

whereby there is induced a sufficient flow
of said fluid upwardly along the wall of
said mixing chamber to (1) entrain solid
particles and maintain particles in the
fluid entrained and (2) inhibit the
build-up of solid particulate polymer
product on the wall of said mixing chamber;
the accumulation of liquid in said mixing
chamber is also inhibited; the upward flow
of said fluid along a central axis of said
mixing chamber being sufficient to prevent
solid particulate polymer product from
falling into said inlet and the combined
velocity and orientation of the said
streams being sufficient to ensure
generally uniform mixing and distribution
of the combined fluid streams by the time
said combined streams reach the upper
regions of said mixing chamber;

    C. passing said combined streams in a
generally uniform manner through a
generally horizontal gas distributor plate
into a fluidized bed region located above
said mixing chamber and said gas
distributor plate at an overall velocity
sufficient to maintain a bed of solid
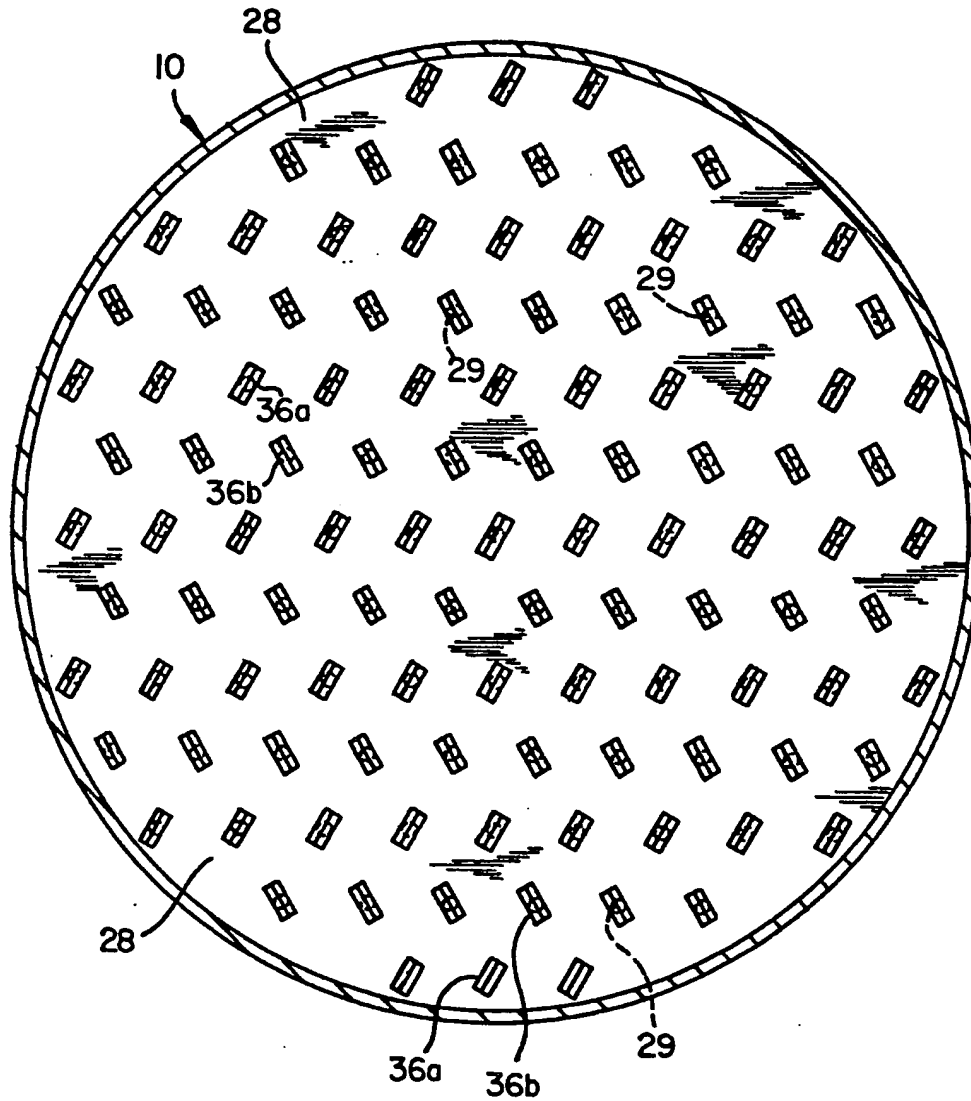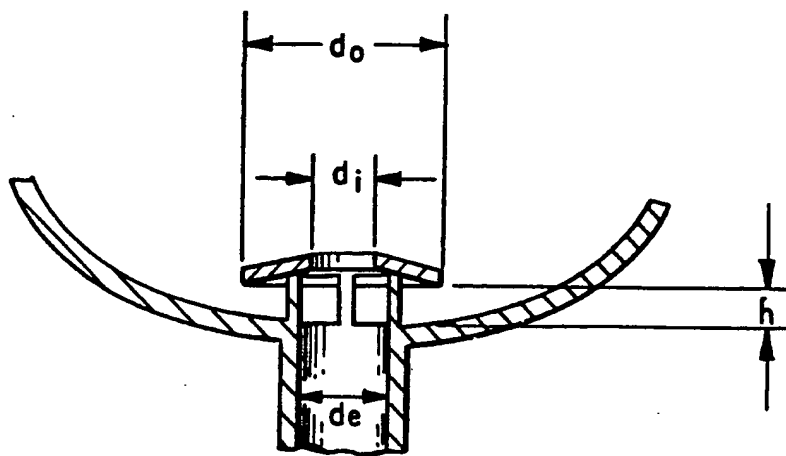polymer particles in a suspended and
fluidized condition.

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FIG. 2

FIG. 2A

FIG. 3
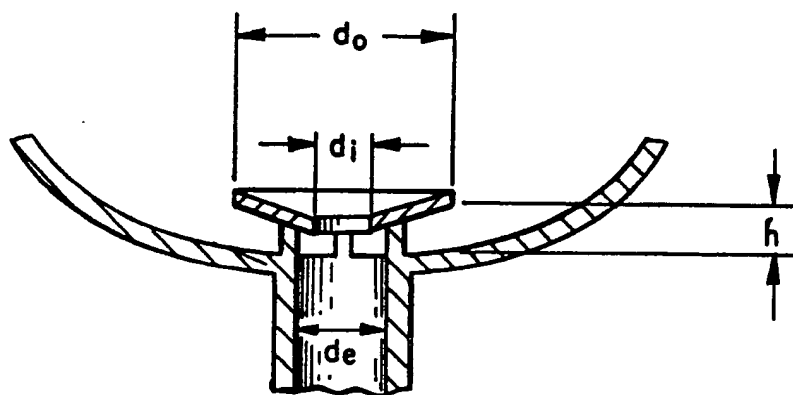
FIG. 4

F I G. 5

F I G. 6